

PROCESS FOR MAKING A LINEAR ALPHA-OLEFIN OLIGOMER
USING A HEAT EXCHANGER

Field of the Invention

The invention pertains to a process for making a linear alpha-olefin oligomer from ethylene in a reactor having a liquid and a gas phase, in the presence of a catalyst.

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Background of the Invention

Various processes are known for the production of higher linear alpha olefins (See, for example, D. Vogt, Oligomerisation of ethylene to higher a-olefins in 10 Applied Homogeneous Catalysis with Organometallic Compounds, Ed. B. Cornils, W.A. Herrmann, 2nd Edition, Vol. 1, Ch. 2.3.1.3, page 240-253, Wiley-VCH 2002). These commercial processes afford either a Poisson or Schulz-Flory oligomer product distribution. In such a process, a 15 wide range of oligomers is typically made.

In WO 02/00339, WO 02/12151, WO 02/06192, WO 02/28805, WO 01/58874, and WO 99/02472 novel Fe-based ethylene oligomerization catalysts are described that show high activity and high selectivity towards linear 20 alpha-olefins. These catalysts, which are incorporated by reference, are based on iron complexes of a selected 2,6-pyridinedicarboxaldehyde bisimine or a selected 2,6-diacylpyridine bisimine.

In the present invention the term "bis-(arylimino)-pyridine" is used to describe both classes of ligands. 25 Alpha-olefin oligomers are compounds or a mixture of compounds with the general formula $H_2C=CH-(CH_2CH_2)_nH$ wherein n is an integer of 1 or greater. In such oligomers the alpha-olefin oligomer is usually a mixture 30 of alpha-olefin oligomers with a mean number n from 1 to

20, preferably from 2 to 10. Alpha-olefin oligomers prepared according to the process of the present invention preferably have an average molecular weight between about 50 and about 350, more preferably between 5 about 60 and about 280, even more preferably between about 80 and about 210.

The reaction of ethylene in the presence of the above iron complex is usually run in the liquid phase in a well-mixed reactor, typically using an aprotic organic 10 solvent. This reaction generates a large amount of heat, which should be removed. As described in WO 02/06192 it is preferred to install a plurality of small reactors in combination with several heat exchangers to help provide sufficient cooling capacity for the reactor system. The 15 process temperature, which usually is between about 35°C and about 90°C, more preferably between about 35°C and about 75°C, affects the cost of manufacture of the alpha-olefins in several ways. The higher the temperature the smaller the heat exchangers which have to be applied to 20 the reactor(s), which generally lowers cost. The decay of the active oligomerization catalyst increases with increasing temperature. It is found that maximum volumetric production of alpha-olefins coupled with good absolute productivity of the catalyst usually occurs in 25 the range of about 45°C to about 75°C, so this temperature range is preferred. Finally, the temperature also affects the bubble point pressure, the amount of ethylene in the liquid phase, and the catalyst 30 selectivity. The higher the temperature the higher the pressure needed to maintain catalyst selectivity, which increases capital cost of the manufacturing plant because of, for example, the need for thicker vessels and larger compressors to attain higher ethylene pressures. Higher pressure also increases energy costs.

The amount of ethylene (ethene) oligomerization catalyst used in the reaction will preferably be the maximum permitted by the cooling capacity of the reactor(s) and the ethylene mass transfer from the gas to 5 the liquid phase. Catalyst may be added to the first reactor only or to one or more subsequent reactors in series. Differing amounts of catalyst may be added to each reactor. The oligomerization is quite exothermic, about 100 kJ/mole of ethylene oligomerized, and as such 10 cooling will usually be applied to the reactor(s) to maintain the desired process temperature while maintaining high volumetric productivity of the reactor(s).

In the prior art cooling is accomplished by running 15 cooling tubes through the liquid in the interior of one or more of the reactors to cool the contents. Another method of cooling is to have one or more heat exchangers external to the reactors and connected to the reactors by a liquid loop to cool the reactor contents. These 20 external heat exchangers may be typical shell and tube exchangers. The reactors may also be jacketed with a cooling jacket. Some or all of the feeds to some or all of the reactors may be cooled to allow the sensible heat of the ingredients to cool the reactors. All these liquid 25 cooling methods, however, suffer from the disadvantage of wax and polyethylene fouling of the coolers, which necessitates regular shut down of the reactor to allow cleaning of the coolers. Furthermore, wax and 30 polyethylene fouling may increase the paraffinicity of the solvent.

Summary of the Invention

It would be advantageous to devise a process without the above disadvantages. It has now been found that linear alpha-olefin oligomers can be made in a reactor

comprising a liquid and a gas phase, comprising the steps of catalytically oligomerizing ethylene in the presence of an iron complex of a 2,6-bis(arylimino)pyridine derivative, to an alpha-olefin oligomer with an average 5 molecular weight between about 50 and about 350 under release of heat, and removing the heat with a heat exchanger, which is not in direct contact with the liquid phase, using at least part of the gas phase as a coolant medium.

10 This method provides a cooling system having its cooling elements outside the liquid reaction medium. Since wax and polyethylene have high boiling points, deposit of wax and polyethylene can no longer occur, and fouling of the heat exchanger is effectively prevented.

15 Brief Description of the Drawings

The invention is illustrated by the following Figures, which are not meant to limit the invention in any way, showing a scheme of an apparatus that can be used for performing the process of the invention.

20 Fig. 1 is a scheme of an apparatus for performing the method according to the invention with the heat exchanger positioned outside the reactor.

25 Fig. 2 is a scheme of an apparatus for performing the method according to the invention with the heat exchanger positioned inside the reactor.

Detailed Description of the Invention

30 The heat exchanger according to this invention is a conventional type, such as a shell- and tube-type, and the like. The heat exchanger is internally cooled with conventional cooling fluids, like water, ammonia, Freon® coolant, and the like. The reaction heat causes the solvents, reactants, and/or reaction products, which are present in the reaction medium, to evaporate and subsequently to be cooled by the heat exchanger, after

which it works as a coolant medium for the reactor. The heat exchanger can be placed inside or outside the reactor. When the heat exchanger is placed inside the reactor it is preferred that some condensation occurs on 5 the heat exchanger surface. When the heat exchanger is placed outside the reactor, it is preferred to apply a forced circulation of the reactor coolant medium from the gas phase of the reactor through heat exchanger(s), compressor(s)/pump(s) and optionally a gas-liquid 10 separator back to the liquid phase of the reactor. Additionally, this will improve the mixing in the reactor. After cooling the reactor coolant medium in this loop, some condensation can occur. This allows application of a separate gas and liquid return to the 15 reactor using a gas-liquid separator. Furthermore, it is possible to deliberately remove (part of) this liquid phase from this gas-liquid separator and route this directly to the product work-up section. Finally, if full condensation occurs, return of this liquid to the reactor 20 can be achieved by a pump instead of a compressor, which lowers costs. This reactor coolant medium is selected from an alkane, alkene, and aromatic compound, and mixtures thereof, preferably propane, n-pentane, isopentane, ethylene, 1-butene, o-, m-, and p-xylene, and 25 toluene, and mixtures thereof.

An additional advantage of the present process is the possibility to apply only one reactor, because the efficiency and the lack of fouling no longer necessitates the use of a plurality of small reactors. This adds 30 considerably to the lowering of costs of the oligomerization process.

The iron complexes of the 2,6-bis(arylimino)pyridine type that can be used in the above process are known in the art, and are described in WO 02/00339, WO 02/12151,

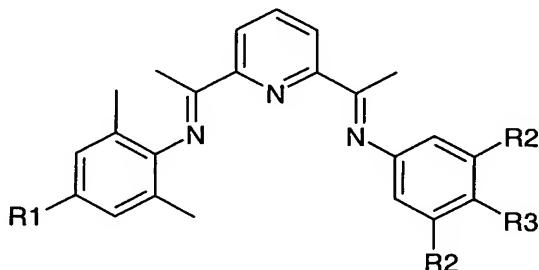
WO 02/06192, WO 02/28805, WO 01/58874, and WO 99/02472.

Any of these complexes can be used. Best results, however, are obtained with such iron complexes wherein one of the aryl moieties of the 2,6-

5 bis(arylimino)pyridine derivative is 2,6-disubstituted with the group CH_2R or $\text{C}_2\text{H}_5\text{R}$, wherein R is selected from H, F, and substituted or unsubstituted aryl, preferably selected from H and F, and the other aryl moiety is 2,6-unsubstituted, or wherein both aryl moieties of the 2,6-
10 bis(arylimino)pyridine derivative are 2,6-disubstituted with F or Cl.

Particularly useful are the 2,6-bis(arylimino)pyridine derivatives with the formula:

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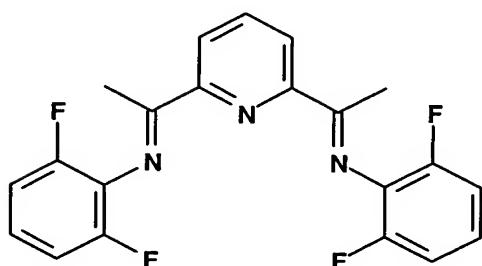
wherein

R1 is H or CH_3 ;

R2 is H, tert-butyl or phenyl; and

R3 is H, tert-butyl or OR' wherein R' stands for CH_3 ,

25 $\text{Si}(\text{CH}_3)_3$ or eicosyl ($\text{C}_{20}\text{H}_{41}$); and



The term "aryl" means an aromatic group, such as phenyl, naphthyl, thienyl, pyridyl, pyrrolyl, and the like.

Phenyl is the preferred aryl group. Preferred phenyl groups are substituted with CH₃, tert-butyl, F, or OR' 5 wherein R' stands for CH₃ or Si(CH₃)₃.

In a preferred embodiment an aluminum-based co-catalyst, preferably a methylaluminoxane, is added to the liquid phase. Where a co-catalyst such as an alkylaluminum compound is required or preferred for the 10 active catalyst species, an iron complex of a 2,6-bis(arylimino)pyridine derivative, such as a complex of the 2,6-bis(arylimino)pyridine derivative with FeCl₂, may be reacted with an alkylaluminum compound, preferably an aluminoxane, to form an active ethylene oligomerization 15 species. Specific alkylaluminum compounds include methylaluminoxane (which is an oligomer with the general formula (MeAlO)_n), (C₂H₅)₂AlCl, C₂H₅AlCl₂, (C₂H₅)₃Al and ((CH₃)₂CHCH₂)₃Al. A particularly preferred aluminoxane is methyl aluminoxane. The ratio of aluminum (as 20 alkylaluminum compound) to iron (as a complex) in the oligomerization may be about 10 to about 10,000.

Another preferred component of the catalyst systems herein is a second co-catalyst compound selected from formula ZnR'₂ wherein each R', which may be the same or 25 different, is selected from hydrogen, optionally substituted C₁-C₂₀ hydrocarbyl, phenyl, F, Cl, Br, I, SR'', NR''₂, OH, OR'', CN, NC wherein R'', which within the same molecule may be the same or different, is C₁-C₂₀ hydrocarbyl.

30 In preferred catalyst systems herein, the second co-catalyst compound is ZnR'₂ wherein R' is C₁-C₂₀ hydrocarbyl, more preferably C₁-C₂₀ alkyl, even more preferably C₁-C₆ alkyl. Suitable alkyl groups include

methyl, ethyl, propyl, butyl, and the like. It is especially preferred that the R' group is a C₁-C₃ alkyl, especially ethyl.

5 The second co-catalyst is particularly valuable in combination with the aluminium-based co-catalyst for increasing the selectivity of linear alpha olefins in ethylene oligomerization reactions, and decreasing the amount of unwanted by-products such as branched olefins, internal olefins, 2,2-disubstituted olefins, and dienes.

10 It has been noted that particularly high selectivity of linear alpha olefins is achieved when the molar ratio of the metal of the aluminium-based co-catalyst to the metal of the second co-catalyst is in the range of from about 5:1 to about 1:5, preferably from about 3:1 to 15 about 1:3, more preferably from about 2:1 to about 1:2 and especially about 1:1.

It is possible to add further optional components to the catalyst systems herein, for example, Lewis acids and bases such as those disclosed in WO02/28805.

20 The active catalyst system may be formed by mixing together the iron complex of a 2,6-bis(arylimino)pyridine derivative or a mixture of the iron acetylacetonate complex and the appropriate 2,6-bis(arylimino)pyridine derivative (ligand), first co-catalyst compound, second 25 co-catalyst compound and any optional additional compounds, preferably in a solvent.

30 An important item in the capital cost of this manufacturing plant and in its cost of operation is the amount of reactor coolant medium that must be recycled in the process. Recycling of a gaseous reactor coolant medium often involves recompression to feed one or more of the reactors. Compressors and associated equipment add greatly to capital and operational costs. In the present method the coolant medium is preferably selected to

completely dissolve ethylene. In this case the coolant medium only requires a single reactor and a condenser, whereas a simple recycle pump is sufficient. Thus expensive recycling, such as the use of an expensive 5 recycle blower, is no longer required, which adds further to the advantages of the present method.

Fig. 1 shows a reactor 2 with a liquid phase 3 and a gas phase 4 being in equilibrium through gas/liquid interface 12. The liquid phase comprises ethylene, the 10 nickel, palladium, cobalt, titanium, zirconium, hafnium, vanadium, chromium, molybdenum, or tungsten complex of a 2,6-bis(arylimino)pyridine derivative, alpha-olefin oligomer, and optionally solvents and auxiliaries such as a co-catalyst. The optional solvents are selected as to 15 dissolve ethylene. The reactor 2 contains an inlet 10 through which the reactor feed 1 (usually ethylene) is introduced into the reactor 2, a gas outlet 11, and a reactor bottom outlet 9. In the embodiment of Figure 1, outlet is connected through a conduit 14 to heat 20 exchanger 5a, which is connected through conduit 15 to gas-liquid separator 6. If necessary, conduit 15 may contain a compressor 7a. Gas-liquid separator 6 has an outlet for transporting the liquid, optionally through a pump 8, to obtain a pressurized liquid stream 17 that is 25 recycled via conduit 19 to reactor 2. The gas leaves the gas-liquid separator 6 through conduit 16, which may optionally comprise compressor 7b and/or heat exchanger 5b, to obtain a cooled gas stream 18 that is recycled to reactor 2. If no condensation occurs in conduit 15, gas- 30 liquid separator 6, and pump 8 are redundant and may be deleted. In that case conduit 15 can directly be connected to compressor 7b and/or heat exchanger 5b, if present, or to conduit 19. Reactor 2 may contain an optional entrainment separator 13.

Fig. 2 shows another embodiment of the invention. In this embodiment the reactor feed 1 is introduced into the reactor 2 through inlet 10. The liquid phase 3 in the reactor is in equilibrium with the gas phase 4 through gas/liquid interface 12. In the section of the reactor containing the gas phase 4, a heat exchanger 20 is placed, which is not in contact with the liquid phase 3. The section of the gas phase 4 may optionally contain an entrainment separator 13. The heat exchanger 20 cools the gas, after which at least part of the gas condenses and the cooled condensate falls down from the surface of the heat exchanger 20 into the liquid phase 3, thereby cooling the liquid medium. The reaction product may then be discharged from the reactor through the reactor bottom outlet 9.

Hence, according to a further aspect of the present invention there is provided an apparatus for performing the process of making linear alpha-olefin oligomer described above, comprising a reactor, which can accommodate a liquid and a gas phase, an inlet through which the reactor feed can be introduced into the reactor, a reactor bottom outlet to remove the oligomer, and a heat exchanger, which is positioned in the gas phase to condense the gas and allow the condensate to fall therefrom to cool the liquid phase, and optionally, an entrainment separator, and/or a gas-liquid separator.